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Chapter 1

Process Intensification: An Overview of Principles and Practice

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1. Introduction

The beginning of the 21st century has been markedly characterized by increased environmental awareness and pressure from legislators to curb emissions and improve energy efficiency by adopting “greener technologies”. In this context, the need for the chemical industry to develop processes which are more sustainable or eco-efficient has never been so vital. The successful delivery of green, sustainable chemical technologies at industrial scale will inevitably require the development of innovative processing and engineering technologies which can transform industrial processes in a more fundamental and radical fashion. In bioprocessing, for example, genetic engineering of micro-organisms will obviously play a major part in the efficient use of biomass but development of novel reactor and separation technologies giving high reactor productivity and ultimately high purity products will be equally important for commercial success. Process Intensification (PI) can provide such sought-after innovation of equipment design and processing to enhance process efficiency.

2. Process Intensification: Definition and Concept

Process intensification (PI) aims to make dramatic reductions in plant volume of between, ideally, 100-1000 fold, by replacing the traditional unit operations with novel, usually very compact designs, often by combining two or more traditional operations in one hybrid unit.

The PI concept was first established in ICI during the late 1970s when the primary goal was to reduce the capital cost of a production system. Although cost reduction was the original target for PI, it quickly became apparent that there were other important benefits to be gained, particularly in respect of improved intrinsic safety and reduced environmental impact and energy consumption, as will be discussed later in this chapter.

Over the last two decades, the definition of PI has thus evolved from the simplistic statement of “*the physical miniaturisation of process equipment while retaining throughput and performance*” [1] to the all encompassing definition “*the development of innovative apparatus and techniques that offer drastic improvements in chemical manufacturing and processing, substantially decreasing equipment volume, energy consumption, or waste formation, and ultimately leading to cheaper, safer, sustainable technologies.*” [2]. Several other definitions with slight variations on the generic theme of innovative technologies for greater efficiency have since emerged [3].

The reduction in scale implied by intensification has many desirable consequences for chemical engineering operations. Firstly, lower mass and heat transfer resistances enabled by the reduced path lengths of the diffusion/conduction interfaces coupled with more intense fluid dynamics in active enhancement equipment allow reactions to proceed at their inherent rates. By the same token, more rapid mixing environment afforded by the low reaction volumes should enable conversion and selectivity to be maximized. Residence times of the order of minutes and seconds may be substituted for the hour-scale processing times associated with large conventional batch operations, with beneficial consequences for energy consumption and process safety.

PI covers a wide range of processing equipment and methodologies as aptly illustrated in Figure 1 [2]. Many of the equipment classed as intensified technologies have been long been implemented in the chemical industry e.g. compact heat exchangers, structured packed

columns, static mixers. More recent developments include spinning disc reactor, oscillating flow reactor, loop reactors, spinning tube in tube reactor, heat exchange reactor, microchannel reactors, etc. Lately, it has become increasingly important for the chemical process industries not only to remain cost competitive but to do so in an environmentally – friendly or “green” manner. It is fitting, therefore, that many of the processes based on the PI philosophy also enables *clean technology* to be practised. For instance, high selectivity operations in intensified reactors on their own will reduce or ideally eliminate the formation of unwanted by-products. Combining such intensified reactors with renewable energy sources such as solar energy would give even greater impetus to achieving these green processing targets.

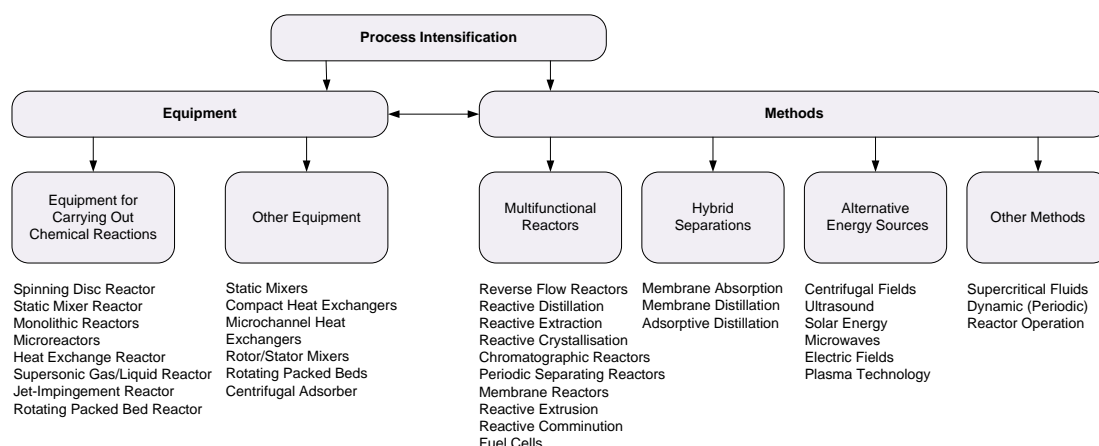


Figure 1. Classification of process intensification equipment and methods [2]

3. Fundamentals of chemical engineering operations

3.1. Reaction Engineering (APH)

Reactor engineering starts with the simple mass balance:

$$\text{In} + \text{Made} = \text{Out} + \text{Accumulated}$$

Where “Made” is the rate at which a species is created or lost by reaction. The rate of this in a well-mixed system depends upon the reaction kinetics, which depend only upon the concentrations of species and temperature. However, not all systems are well-mixed, particularly at larger scales, and mixing can be rate determining. The different degrees and types of mixing are introduced in section 3.2. The “accumulated” term will be zero for continuous reactors running in steady state, but will be of interest during start-up or shut-down. Determining the rate at which species are created or destroyed in a reactor requires knowledge of mixing, reaction kinetics and heat transfer. Once these are known they can be input into a reactor model. An important part of this model for continuous reactors (as most intensified reactors are) is the Residence Time Distribution (RTD), which is the probability distribution for a given reactor design the length of time elements of fluid spend in the reactor. It can be envisaged as the response to the input of an infinitely narrow pulse of a tracer. All real reactors fall between two extreme cases: the Plug Flow Reactor (PFR) and the Continuously Stirred Tank Reactor (CSTR).

3.1.1. The Plug Flow Reactor (PFR)

“Plug Flow” refers to fluid flowing in discrete “plugs” i.e. without interaction between elements of fluid. The RTD of a perfect plug flow reactor is infinitely thin. Any input tracer pulse to the reactor will remain unchanged as shown in Figure 2.



Figure 2. A perfect Plug Flow Reactor (PFR), showing the response to a perfect input pulse

Real PFRs have symmetrical Gaussian RTDs centred on the mean residence time, the breadth of the RTD decreasing with increasing proximity to ideal plug flow. In practice, this is usually achieved by ensuring a high level of turbulence in the flow, as it produces a flat velocity profile. The most conventional form of PFR is a tubular reactor in very turbulent flow. However, there are many variations on this basic form, and other ways of achieving plug flow. Chapters 3 and 5 cover examples of unconventional, intensified plug flow reactors.

3.1.2. The Continuously Stirred Tank Reactor (CSTR)

The CSTR is, at its simplest, a batch stirred tank to which an inflow and outflow have been added (of equal flowrate, when at steady state). To determine the RTD of such a reactor we must picture a pulse of fluid entering it. A “perfect” CSTR is perfectly mixed, meaning that fluid is uniformly dispersed the instant it enters the reactor. The outflow is at the same concentration of tracer as the bulk of the reactor. Tracer will begin to flow out at this concentration (initially), whilst being replaced with fluid containing no tracer i.e. the tank gradually become diluted, and the concentration in the outflow will decrease. This leads to a monotonic decrease in concentration, which can be shown to follow an exponential decay (Figure 3).

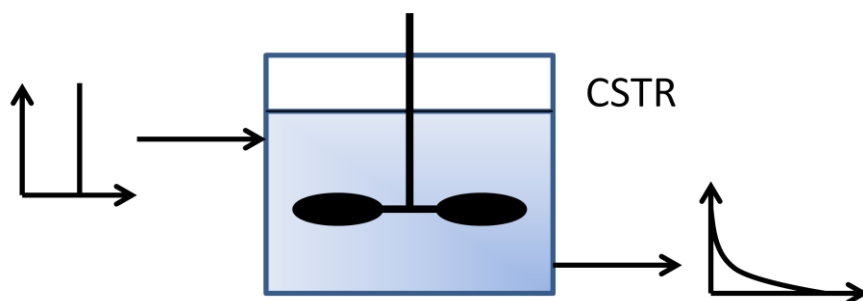


Figure 3. RTD for an Ideal CSTR

3.1.3. The Plug Flow Advantage

The CSTR's RTD is generally not desirable, as, for a given desired mean average residence time:

- i. much of the material in the reactor will spend too long in the reactor (due to the long tail in the RTD) and will consequently be “overcooked”. The main problem is that this allows competing reactions to become more significant.
- ii. much of the material will be in the reactor for less than the desired residence time. This will not therefore reach the desired level of conversion

The CSTR can therefore lead to increased by-product formation and unsatisfactory conversion. In contrast, plug flow means that each element of fluid experiences the same processing history: each spends exactly the same amount of time in the reactor as any other, and is subject to exactly the same sequence of conditions. This reduces byproduct formation, and ensures that the desired conversion is achieved. Furthermore, in practice, this will reduce the size of the reactor compared to a CSTR for the following reasons:

1. The reactor will be the correct size. CSTRs are usually oversized to compensate for the poor RTD.
2. No headspace is required, as is the case in any tank reactor.
3. For most reaction kinetics (the most notable exception perhaps being autocatalytic reactions) simply following the design equations will lead to a PFR design that is smaller than a CSTR. For an explanation of this, the reader is advised to consult sections 5.2 and 5.3 in reference [4].
4. Stirred tanks do not scale up in a predictable manner. Uniform mixing becomes difficult to achieve, which can reduce the rate of reaction, necessitating a larger reactor. This is less of an issue with tubular reactors.

For these reasons, PFRs are often preferred in principle. In practice they are difficult to use at long residence times (above a few minutes) and multiphase reactions can be difficult.

3.2. Mixing Principles

Mixing is the process of bringing in close proximity separated fluid elements in a system which, in the simplest case, aims to reduce non-uniformity in a particular property, for example, concentration, viscosity or temperature. Most mixing processes occur alongside heat and/or mass transfer operations and chemical reactions.

3.2.1. Influence of mixing on reactions

Mixing is a particularly important process in reactor design, especially in continuous flow reactors. Designing the mixing process to yield a much shorter mixing time in comparison to the mean residence time of the reactants in the reaction vessel is of paramount importance for good operation of the reactor. If mixing is slow, large and varying concentration gradients of reactant species will exist in different parts of the reactor, resulting in wide variations in product concentrations and properties, which may be deemed off-spec in many applications. In fact, the rate of mixing often determines the rate of these processes and may have a significant impact on the product distribution obtained, especially if many competing reaction steps are involved.

3.2.2. Turbulent mixing: mixing scales, mechanisms and mixing times

In a single-phase turbulent flow system, there are three distinct mixing scales that influence a chemical process: macromixing, mesomixing and micromixing [5,6]. They are defined on the basis of their characteristic length scale, as depicted in Figure 4, and are directly correlated with the turbulent energy dissipation rate, ϵ .

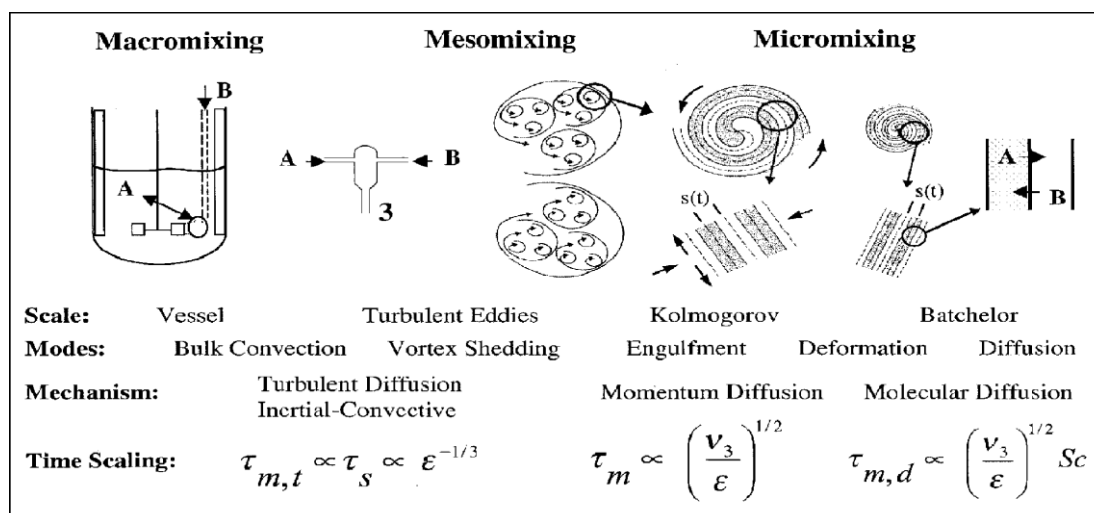


Figure 4. Turbulent mixing mechanisms across various length scales [7]

The intensity of mixing at each of the scales referred to above is significantly influenced by the mechanical energy input into the system by the mixing device. It is generally assumed that higher energy input translates into higher energy dissipation rate for better mixing- but this is not always the case, as energy may be wasted, for example, in vortex formation at higher agitation rate in an unbaffled vessel. The energy input causes the fluid to undergo motion across the cascade of length scales described below so that any concentration inhomogeneities is gradually reduced and eliminated. The kinetic energy thus imparted to the fluid is ultimately dissipated as internal energy, which occurs at the smallest length scales of turbulence i.e at the Kolgomorov scale.

Various mixers/reactors have been characterized in terms of their energy dissipation rates, as shown in Table 1. This illustrates the potential capability of intensified systems such as static mixers, rotor-stator mixers and the SDR amongst others to provide a higher level of mixing intensity than the conventional stirred tank reactor. It is important to remember, though, that higher energy input will be a penalty to incur in term of energy consumption and the benefits from the mixing process under these conditions have to demonstrate significant process improvement.

Table 1. Comparison of energy dissipation rates in a range of mixers/reactors

Reactor/mixer type	Energy dissipation rate (W/kg)	References
Stirred tank reactor	0.1-100	[8,9]
Static mixers	1-1000 W/kg	[9,10]
Impinging jet reactor	20-6800 W/kg	[11]
Rotor-stator spinning disc reactor (27 cm disc diameter, 240-2000 rpm)	Up to 6000 W/kg	[12]
Thin-film spinning disc reactor (10 cm disc diameter, range of disc speeds 200-2400 rpm)	Up to 2000 W/kg	[13]

3.2.2.1 Macromixing

Macromixing involves mixing on the macroscopic scale, which refers to the scale of the vessel or reactor. The process is often referred to as distributive mixing [6,14] which is achieved by bulk motion or convective transport of the liquid at the macroscopic scale, resulting in uniform spatial distribution of fluid elements within the reactor volume. In a continuous flow reactor, the macromixing process directly influences the residence time distribution of a feed stream introduced into the contents of the vessel.

The macromixing time in a mechanically stirred, baffled tank, τ_{mac} , is a function of the mean circulation time, τ_c , in the vessel. In a vessel configured for optimized mixing, $\tau_{\text{mac}}=3\tau_c$, whilst in a non-optimised system, $\tau_{\text{mac}}=5\tau_c$ [6].

The mean circulation time, τ_c , is generally expressed in terms of the impeller pumping capacity Q_c [14]:

$$\tau_c = \frac{V}{Q_c}$$

Or

$$\tau_c = \frac{V}{C_D N D_i^3}$$

where C_D , the discharge coefficient of the impeller, is a constant which varies typically between 0.7 and 1.0 depending on the impeller used [14].

3.2.2.2 Mesomixing

Mesomixing refers to coarse-scale, dispersive mixing via turbulent eddies. It is typically characterized by two different mechanisms [5,15]: (1) turbulent dispersion of a fresh feed introduced to a vessel which mixes with its local surroundings (2) inertial- convective break-up of large eddies which are larger than the Kolgomorov length scale defined below.

Turbulent dispersion characteristic timescale, τ_D , can be defined by either equation below, depending on the radius of the feed pipe, r_{pipe} , with respect to the characteristic length scale for dispersion, L_D [5,15]:

$$\tau_D = \frac{Q_f}{u D_{\text{turb}}} \quad (\text{if } r_{\text{pipe}} \ll L_D)$$

$$\tau_D = \frac{r_{\text{pipe}}^2}{D_{\text{turb}}} \quad (\text{if } r_{\text{pipe}} \approx L_D \text{ or } r_{\text{pipe}} > L_D)$$

where $D_{\text{turb}} = 0.12 \varepsilon^{1/3} L_D^{4/3}$ [5,16]

Baldyga et al. [15] have expressed the inertial-convective mesomixing timescale, τ_s , as:

$$\tau_s = A \left(\frac{L_c^2}{\varepsilon} \right)^{\frac{1}{3}}$$

where A is a constant having a value between 1 and 2, depending on the turbulence level in the system.

3.2.2.3 Micromixing

Micromixing represents the final stage of the turbulent mixing process, which proceeds at much finer length scales than macro and meso-mixing, referred to as the Kolgomorov or Batchelor length scales. At the microscale level, the Kolgomorov length scale, η_K (representing smallest scales of *turbulence* before viscosity effects dominate) and Batchelor length scale, η_B (representing smallest scales of fluctuations prior to *molecular diffusion*) are defined as [17,18]:

$$\eta_K = \left(\frac{\nu^3}{\varepsilon} \right)^{1/4}$$

$$\eta_B = \left(\frac{\nu D_l^2}{\varepsilon} \right)^{1/4} = \frac{\eta_K}{\sqrt{Sc}}$$

where the Schmidt number, $Sc \left(= \frac{\nu}{D_l} \right)$ for liquids is typically of the order of 10^3 , so that $\eta_B \ll \eta_K$. For aqueous solutions in turbulent regimes, η_K is of the order of 10-30 μm .

The physical phenomena of the micromixing process include engulfment, deformation by shear and diffusion of the fine scale fluid elements. The relevant mixing times associated with these processes are [5]:

Engulfment:
$$\tau_e = 17.2 \left(\frac{\nu}{\varepsilon} \right)^{0.5}$$

Shear deformation and diffusion:
$$\tau_{Ds} \approx 2 \left(\frac{\nu}{\varepsilon} \right)^{0.5} \text{arc sinh } (0.05 Sc)$$

More often than not, $\tau_{DS} \ll \tau_e$, resulting in the overall micromixing process being dictated by the progression of the engulfment phenomenon which takes place at the Kolgomorov length scale.

Although the actual molecular mass transfer process before reaction takes place is ultimately achieved by molecular diffusion, enhancing the rates of macro and mesomixing through turbulent hydrodynamic conditions enables faster attainment of the fluid state where micromixing and therefore molecular diffusion prevails.

3.3. Transport processes

Understanding transport processes is at the heart of process intensification, as the subject can be defined as a search for new ways of enhancing or achieving transport of mass, heat or momentum.

Transport processes, i.e. heat, mass and momentum transfer, are generally governed by equations of the same form. They are all flows in response to a “driving force”, a temperature difference, a concentration difference or a pressure difference, respectively, opposed by their respective resistances.

3.3.1 Heat Transfer

Heat transfer – the transport of energy from one region to another driven by a temperature difference between the two – is a key consideration in the design of all unit operations. All unit operations have defined operating temperatures, so the heat flows in and out must be understood, to maintain the temperature within a desired range. Reactors, for instance, must be supplied with heat or have it removed at a rate depending upon the exo/endothermicity of

the reaction, the heat transfer characteristics of the reactor and the heat flows in and out, to ensure that the reaction takes place at the correct temperature and therefore rate.

Furthermore, the streams into and out of unit operations must be maintained at the correct temperatures. This is usually achieved using heat exchangers, devices for transferring heat between fluid streams, without the streams mixing. It had always been a given in heat exchanger design that they must operate in turbulent flow wherever possible, as turbulent flow results in considerably higher heat transfer coefficients than laminar. Hence, heat exchangers were not designed with narrow channels, as the achievement of turbulence depends upon exceeding a certain Reynolds number, which is directly proportional to the diameter of the channel:

$$Re = \rho v D / \mu$$

Reassessing such assumptions about heat ^{*Re*} and mass transfer is at the heart of process intensification, and here led to the the development of “compact heat exchangers”, which have extremely narrow channels.

This only makes sense if the heat transfer rather than just the heat transfer coefficient is considered. The rate of heat transfer in a heat exchanger is not only a function of the heat transfer coefficient, as can be observed in the “Heat Exchanger Design Equation” below:

$$q = UA\Delta T_{lm}$$

It is also clearly a function of the heat transfer surface area. Compact heat exchangers have very narrow channels (sub-mm), so the flow is laminar (as Re depends upon channel width, D) and therefore has a significantly lower heat transfer coefficient than a turbulent flow. However, this is more than compensated for by the increase in heat transfer surface area per unit volume, giving a higher heat transfer rate per unit volume than conventional heat exchanger designs (such as "shell-and-tube"). A concise overview of compact heat exchangers is given by Reay et al. [19].

There are also a range of devices ("turbulence promoters") that are designed to perturb the flow to bring about the onset of turbulence at lower Re . The promoters allow the higher heat and mass transfer coefficients associated with turbulence to be accessed at lower velocities, thereby reducing the associated pumping duties. These can also be classified as intensified devices, although the degree of intensification is nowhere near as great as in the compact heat exchanger. They, however, suffer less from fouling, which is one of the main drawbacks of compact heat exchangers: their applications are limited to "clean" fluids, as they are very easily blocked by fouling. As with most technologies, the strengths and weaknesses of intensified technologies must be assessed, so as to define a "niche" or parameter space within which they are the best-performing technology.

3.3.2 Mass Transfer

An appreciation of mass transfer is required for the intensification of separation processes. Common separation unit operations are distillation, crystallization, ad/absorption and drying.

In many processes of course the heat and mass transfer are interrelated. Generally what enhances one enhances the other. Indeed, the mechanisms for transfer are often the same or closely related. Experiments in heat transfer have often been used to draw conclusions about mass transfer (and vice versa) by various analogies. Various equations describing one or the other are based upon analogy. Compare for instance the Dittus-Boelter equations for heat and mass transfer:

$$\text{Heat: } Nu = C_1 \cdot Re^{0.8} \cdot Pr^{0.33}$$

$$\text{Mass: } Sh = C_2 \cdot Re^{0.8} \cdot Sc^{0.33}$$

An example of an intensified mass transfer device is the rotating liquid-liquid extractor. The conventional design of liquid-liquid extractors was based on using the density difference between the liquids to drive a countercurrent flow, by inputting the denser fluid at the top of the column, and the lighter at the bottom. One of the variables, although it may not appear to be a variable initially is g , the acceleration due to gravity. g can of course be increased by applying a centrifugal field, in which case the lighter fluid is introduced from the outside and travels inward countercurrent to the denser fluid. The first example of this kind of device is the Podbielniak liquid-liquid contactor originally developed in the 1940s for penicillin extraction. There are hundreds of Podbielniak contactors in use worldwide currently for a wide range of applications including has been used for many applications, including antibiotic extraction, vitamins refining, uranium extraction, removal of aromatics, ion exchange, soap manufacture and extraction of various organics [20]. This illustrates that there are many successful examples of process intensification in industry today, although they are not viewed as such, as they are not new technology (and the term process intensification did not exist when they were invented). Indeed, any continuous process is an example of an intensified process.

4. Intensification techniques

Intensification of a process may be achieved through a variety of ways including enhancing mixing and heat/mass transfer by additional energy input via external force fields or via enhanced surface configurations, transforming processes from batch to continuous mode in order to achieve smaller process volume and integrating process steps in hybrid technologies. Each of these will be discussed briefly in the ensuing sections.

4.1. Enhanced Transport Processes

Heat and mass transport rates are largely influenced by the fluid dynamics which directly affect the heat/mass transfer coefficients and the available area for the transfer of energy/mass to occur. Mixing rates are similarly affected by these parameters. Therefore, any attempt at intensifying these processes should focus on enhancing the turbulence in the system and on increasing the transfer surface area. One way of achieving this is by subjecting the reaction environment to external force fields such as centrifugal, electric and ultrasonic fields.

4.1.1. Enhanced Force Fields

4.1.1.1. Centrifugal fields

Surface rotation as a technique for intensification has stimulated keen interest from academic workers for many years. As early as the 1950s, Hickman's research efforts into two phase heat transfer on spinning disc surfaces culminated in the development of the first successful centrifugal evaporator used in sea-water desalination [21].

The benefits that can be extracted from the exploitation of high centrifugal fields created by rotation are outlined below:

1. The rotational speed of the spinning surface provides an additional degree of freedom that can be readily manipulated for optimum equipment performance.

2. The extremely high gravity fields thus generated are capable of producing very thin films in which heat transfer, mass transfer and mixing rates are greatly intensified. The short path lengths and the high surface area per unit volume provide the opportunity for rapid molecular diffusion and enhanced heat transfer, even on scale-up (Figure 5). The performance of multiple phase processes in particular stands more chance of being enhanced under the influence of high gravitational forces as a result of increased inter-phase buoyancy and slip velocity [22].
3. Applications in which the solid content of a process fluid often poses a number of problems with regard to fouling in conventional devices can, in principle, be handled by the rotating equipment. The rotating action in itself provides a scraping or ‘self-cleaning’ mechanism strong enough to shift most solid deposits away from the surface of revolution, thereby ensuring maximum exposed area at all times during operation.
4. The very short and controllable residence times achieved under the centrifugal action enable heat sensitive materials to be processed with minimal risk of degradation.

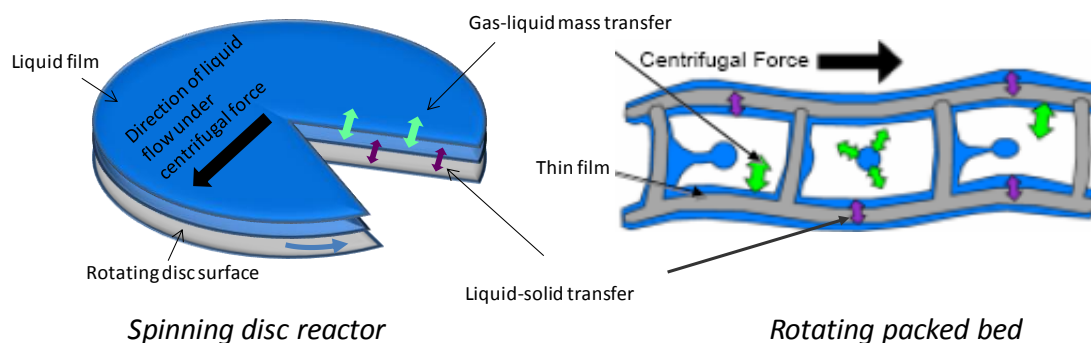


Figure 5. Thin film processing in spinning disc reactor and rotating packed bed, illustrating the short diffusion/conduction path lengths and high surface area for heat transfer

Several unit operations have been identified where centrifugal acceleration generated on surfaces of revolution presents remarkable potentials for intensification. Typical operations

include distillation, extraction, boiling, condensation, crystallisation, precipitation and gas-liquid reactions among others.

The spinning disc reactor and the rotating packed bed are two well-known examples of centrifugal field processing equipment. The SDR will be treated in more detail in Chapter 3 of this book.

4.1.1.2. Alternative force fields

Alternative force fields commonly employed to intensify processes include ultrasound, electric fields, and energy of electromagnetic radiation whose applications to chemical and biochemical processes in the context of process intensification have been recently reviewed by Stankiewicz [23].

4.1.1.2.1. Ultrasonic fields

Ultrasound refers to sound waves beyond the audible range of the human ear, in the frequency range of approximately 20 kHz to 500 MHz. The range typically applied to chemical processing is generally no higher than 2 MHz [24]. Ultrasound is propagated through a liquid medium in alternating cycles of compression and stretching, or rarefaction. These induce an effect known as cavitation whereby microbubbles are generated, expand and are subsequently destroyed in successive compression cycles releasing a large amount of heat and pressure energy in the local environment of the bubbles (Figure 6). Local temperatures and pressures after the collapse of microbubbles can reach as high as 5000°C and 2000 atmospheres respectively depending on the power input [24]. Mechanical effects or chemical effects can arise from such extreme conditions in the system, as discussed in many review articles on the subject [25, 26]. Thus, for instance, the mechanical effects are characterized by the pressure waves or shock waves resulting from the collapse of cavitation bubbles. These waves generate intense mixing conditions and enhanced transport rates

throughout the bulk of the liquid medium in homogeneous systems and at liquid/liquid or liquid/ solid interfaces in heterogeneous systems, which have a direct, positive influence on a chemical reaction. Furthermore, in immiscible liquid systems, much finer droplets can be formed under ultrasound exposure than by mechanical agitation, creating higher surface area for mass transfer. These mechanical or physical effects are generally thought to be responsible for the rate enhancements and improved product properties observed in many chemical processes subjected to ultrasonic irradiation [26, 27].

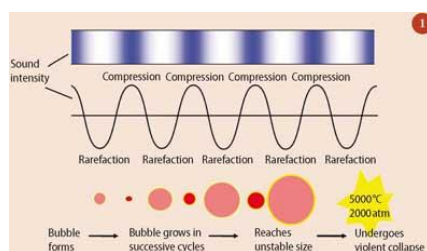


Figure 6. Generation and collapse of an acoustic cavitation bubble [27]

Chemical effects due to ultrasound arise if the chemical compounds in the processing medium can fracture into reactive intermediates (often referred to as sonolysis) such as free radicals under the very intense local temperatures attained in the bubble cavities. The literature has many documented examples of such sonolysis processes [24, 28, 29]. Such chemical effects are usually reflected in changes in mechanisms and product distributions [30]. The well-documented influence of ultrasound on crystallization processes illustrates the latter effect most appropriately [31].

In summary, ultrasonic processing for intensification of chemical processes is associated with the following benefits, all of which have wider implications for greener processing [24, 27]:

- Increases in both reaction speed and yield in an extensive range of heterogeneous and homogeneous systems, as highlighted by Thompson and Doirasawmy [26]

- Waste minimization through increased selectivity
- Changes and simplification of reaction pathways which can lead to milder processing conditions (e.g ambient temperatures and pressures, reduced solvent use) and higher energy efficiency
- Use of environmentally benign reactants and solvents whilst retaining or even enhancing reaction rate under ultrasonication

Chapter 7 explores the above concepts in more detail.

4.1.1.2.2. Electric fields

High intensity electrical fields have long been known to have a destabilizing effect on dispersed systems containing polar molecules such as water and to enhance mass transfer processes via promoted coalescence of the dispersed phase. The removal of dust from air in the Cottrell Precipitator [32] and the dehydration of crude oil emulsion in oilfields were amongst the first industrial processes developed almost a century ago to harness the beneficial effects of electric fields in such phase separations. The essential features of the mechanism involved in such electrostatic separations involve [33]:

- (1) Charging of a liquid droplet- this can be either by (a) induced charging of polar molecules via polarization and reorientation of molecular dipoles charging by the applied electric field, or (b) direct or contact charging of non-polar molecules by contact with a DC charged electrode
- (2) Aggregation, coalescence and settling under gravity of electrically charged droplets to achieve complete phase separation

Intensification of electrical field separations have focused on improved designs of the coalescing vessel and the electrodes, enhanced hydrodynamics to promote electrically charged droplet interactions through turbulence and higher electric field strengths [33]. The large interfacial areas formed due to small droplet formation in electric fields can also be beneficially applied to enhance overall rates of reaction in immiscible liquid systems whereby a higher degree of stable emulsification is achieved [34, 35]. In one such process involving the enzymatic hydrolysis of tri-glyceride esters to yield free fatty acids and glycerol studied by Weatherley and Rooney [36], electrostatic fields have been used to intensify the dispersion of the aqueous phase into the oil substrate by creating large interfacial areas between the reacting species and thus enhance the overall rate of reaction at relatively modest temperatures and pressures.

Electric field effects on intensification of heat transfer processes in general and of boiling in particular, are also well documented [37-40] and the mechanisms are generally well understood [41-43]. In nucleate boiling, for instance, not only are more bubbles released from the surface when an electric field is applied but these are also smaller than in its absence [42].

The combined effects of electric and centrifugal fields on hydrodynamics in thin film flow in a spinning disc reactor have been analysed via numerical simulations by Matar and Lawrence [44]. The applied electric field was shown to induce turbulence on the film surface through the formation of an increased intensity of large amplitude waves. These simulation results suggest that electric fields have the potential to further enhance heat and mass transfer and reaction rates in thin film processing.

4.1.1.2.2. Electromagnetic fields

The electromagnetic spectrum covers a wide range of energy fields such as microwaves, light, X-rays and γ rays. Here only microwaves and light will be considered for the intensification of chemical processes.

Microwaves

Microwaves are a form of electromagnetic energy with frequencies in the range of 300 MHz to 300 GHz. The commonly used frequency for microwave heating and chemical processing is 2.45 GHz. Microwave heating of materials is quite distinctive from conventional heating where conduction and convection are the main mechanisms for the transfer of heat. Under microwave exposure, on the other hand, dipole interactions or ionic conduction come into play depending on the chemical species involved [45, 46]. Dipole interactions occur with polar molecules having high dielectric constants such as water and alcohols whilst migration of dissolved ions in the electric field takes place in ionic conduction. Both mechanisms require effective coupling between components of the target material and the rapidly oscillating electrical field of the microwaves. Heat is generated by molecular collision and friction.

Microwave assisted processing is associated with many features which have positive implications for intensification of processes. In particular, the more rapid, controlled and uniform heating rates afforded by microwave exposure results not only in higher rates of reactions rates than conventional heating methods but also in better product quality through improved selectivity. Recent publications by Toukoniitty et al. [25] and Leonelli and Mason [27] provide good reviews of these aspects, with the former focusing on heterogeneous catalytic systems.

There are a number of examples of industrial use of microwave heating in the food, rubber and wood industries as highlighted by Leonelli and Mason [27]. Although small scale batch and continuous reactor systems with microwave irradiation capabilities are commercially

available [45], there is much scope still for further development of industrial scale microwave reactors, with continuous flow systems having microwave irradiation or indeed coupled microwave and ultrasound irradiation capabilities being ideally suited for this purpose [23,27].

Light energy

The observation that certain compounds could be affected by sunlight to give materials having the same chemical composition but very different physical properties was made as long ago as 1845 when Blyth and Hofmann [47] noted that styrene was converted from a liquid to a glassy solid when exposed to sunlight. Since then, the industrial potential of photochemistry has been widely demonstrated in a number of reactions such as the production of caprolactam used in the manufacture of Nylon 6 and the formation of vitamins D₂ and D₃ [48].

Photo-initiation is an attractive alternative to thermal activation of reactions for a number of reasons. It is an inherently clean process requiring only the reacting molecules for the reaction to be activated. Additional and often expensive and environmentally unfriendly reagents and catalysts can be minimized. The irradiation by a specific wavelength and therefore a well-defined energy input in the form of photons allows only certain reactions to be targeted, thereby reducing by-product formation and costly downstream separation processes. Since activation is by light, ambient operating temperatures can be utilized which not only reduces thermal energy consumption but can also result in better control of the process, especially if side reactions can be minimized at lower temperatures. This is particularly relevant for polymer processing where low temperatures generally result in (1) better tacticity control of the polymer (2) reduced transfer effects, which can be a cause of

excessively branched macromolecules and (3) minimal thermal degradation risks of the polymer formed.

In spite of the clear advantages offered by photochemistry as a reaction initiation technique, its usage in industry is surprisingly rather limited. This is mostly because of the technical problems associated with the uniform radiation of large reaction volumes such as those encountered in conventional batch set-ups. Batch reactors, especially those at commercial scale, possess particularly low surface area to volume ratios. This poses enormous processing challenges in that the photons emanating from the light source have extremely limited penetration depths into the fluid (a few centimeters at most), resulting in quite ineffective and non-uniform initiation of reactions in conventional stirred tank reactor configurations. These issues have been addressed to a certain extent by the development of the falling film reactor but other reactor configurations such as the thin film spinning disc reactor or microreactor could potentially be applied for more efficient processing of photochemical processes. This will be dealt with in more detail in Chapter 3 where the spinning disc reactor will be considered.

4.1.2. Enhanced surface configurations

The influence of well-defined structures or enhanced surface configurations on the increased efficiency of chemical processes, in terms of heat/ mass transfer and mixing rates, is well established. Such structures can be classified according to the time and length scales at which they function. For example, molecular scale structures such as zeolite supports, which are termed “molecular reactors” [6] have been used to improve chemical transformations at the molecular scales. A range of examples of such molecular reactors applied in chemical reactions is provided in a recent paper by Van Gerven and Stankiewicz [6] . At the meso- and macro-scales, structures such as channel reactors (including milli-and micro channels),

monoliths, foams and static mixers have all been used to improve process performances such as yield and selectivity. A brief overview of these areas is given below; a selection of these will form the subject of later chapters in this book.

4.1.2.1 Micro-/milli channel reactors

The use of micrometer or millimetre scale (typical dimensions of 10-100 μm and 0.5-2 mm, respectively) reaction spaces in the form of channels of various shapes (Figure 7) allows for much more precise control of diffusion, heat exchange, retention/residence times and flow patterns in chemical reactions [49]. Although flow in these types of reactor systems tends to be mainly laminar ($\text{Re} < 100$ typically), effective mass transfer can be achieved via the very short diffusion path lengths, determined by the diameter of the channels. Thus, given that the characteristic diffusion time, t_D , is expressed as:

$$t_D = \frac{l^2}{D_l}$$

the diffusion time across a macroscale channel (of the order of 1 cm diameter) and a microchannel of 100 μm diameter can vary dramatically from 10^5 s to 10 s respectively, for solvated molecules of typical diffusivity 10^{-9} m^2/s . Although the diffusion timescale is greatly reduced in the microchannel, it is nevertheless still a slow process for efficient micromixing in fast reactions where micromixing times of the order of milliseconds are required. In such cases, it is therefore important to also provide very large interfacial areas for efficient contact between reacting streams in the channel, which can be achieved through a number of “passive” techniques as described by Hessel et al. [50] e.g. through the use of miniaturised static inserts which continuously split and recombine the flow streams or by turbulent collision of injected streams, amongst many others, albeit at the expense of greater pressure drop. Moreover, when more than one phase is present, vortex motions generated by

the shearing motion within the slug flow are also known to enhance the mixing within the slugs and improve mass transfer rates across the interface [51, 52]. Near plug-flow behaviour giving control of residence time distributions in micro/milli-channel reactors can be achieved by introducing segmented flow via a second phase in the flowing system [50]. Very good control of highly exothermic reactions in small diameter channel reactors is also possible due to the extremely high surface area to volume ratios of up to $50000 \text{ m}^2/\text{m}^3$, when compared with conventional stirred tank reactor geometries.

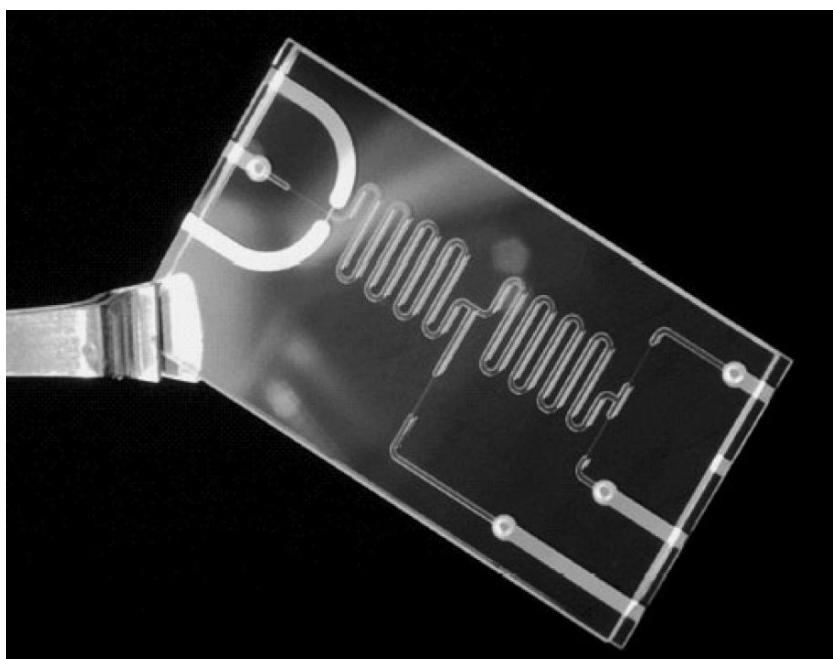


Figure 7. A glass microreactor [53]

As a result of these admirable traits, much attention has been given to micro and meso-scale channel reactors as alternative systems for a host of reactions that are typically limited in some way (mass transfer, heat transfer or mixing) when conducted in conventional batch reactors. Mason et al. [54] offers an extensive review of the current literature in this area but specific examples worthy of note include hydrogenations reactions [55], a nitration reaction [56] and a simple acid-base reaction [51], all of which require rapid mass transfer in order to keep pace with the inherently fast reaction kinetics.

The characteristics of microreactor technologies and their processing advantages will be covered in greater depth in Chapter 4 of this book.

4.1.2.2 Monolithic structures

Monoliths represent a type of structured packing unit consisting of a large number of parallel, straight capillary channels, separated from each other by walls. The “honeycomb” channels come in various shapes and sizes channels with rectangular channels being most commonly used. They are usually made from ceramic or metallic materials and have traditionally been used as catalyst supports in the treatment of NO_x and CO emissions in the automotive industry [57]. Monolithic catalyst structures are characterised by several features which make them highly attractive for use in multiphase catalytic reactors [57, 58]. In brief, flow in the channels is uniformly distributed, with relatively low pressure drops. The large catalyst-coated surfaces areas presented by the monolith arrangement, which can be further augmented by the provision of fins in the channel walls, are conducive to extremely good contact between the gas/liquid phase with the catalyst. Combined with the short diffusion path length offered by the thin catalyst coating, the large contact areas between the catalyst and the reacting phases give rise to intense mass transfer to and from the catalyst.

Other structures such as porous meshes [59, 60] and metal foams [61] have also been shown to enhance gas-liquid mass transfer and uniformity of mixing respectively.

Monolith catalytic reactors will be covered in more detail in Chapter 6.

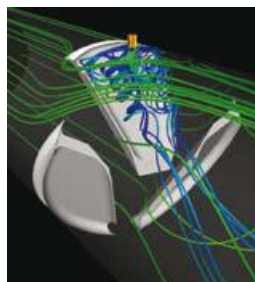
4.1.2.3 Static mixers

Static mixers are motionless pipeline inserts designed to promote mixing. Various static mixer designs are available for a range of applications including blending of viscous

materials and laminar or turbulent mixing in single phase liquids or multiphase systems. For laminar mixing, the mixer elements such as those of the SMXTM configuration (Figure 8a) are designed to repeatedly split the incoming flow into many layers which are redistributed around the mixer structure in the transverse direction relative to the net flow. This gives rise to greater uniformity of component distribution across the flow cross section of the pipe and increased interfacial area for enhanced diffusion. Static mixers designed for turbulent mixing on the other hand seek to enhance the formation of turbulent eddies in the flow stream (Figure 8b).



(a)



(b)

Figure 8. (a) SMXTM static mixer for laminar mixing (b) CompaXTM static mixer for turbulent mixing (Images courtesy of Sulzer Chemtech)

Several benefits of static mixers can be exploited, including homogenous mixing, narrow residence time distribution, low cost and low maintenance due to lack of moving parts. One of their limitations is related to the handling of very viscous materials which may result in extremely high pressure drops, although new designs such as the SMXTM plus mixer have been introduced to alleviate such processing difficulties [62]. Moreover, unless the structures are very open, static mixers can be prone to blockages by suspended solids (e.g. catalysts).

Static mixers have found wide applications in many industrial mixing operations involving gas-liquid dispersion and liquid-liquid extractions for enhanced mass transfer [63] and have long been regarded as one of the successful applications of process intensification in industry. They can also be employed as an in-line mixer in multifunctional heat exchanger-reactors where, not only do they facilitate rapid mixing between reacting species, but can also provide additional surfaces for efficient heat transfer [64-66].

In summary, the heat and mass transfer intensification ability of the various technologies as described in section 4.1 is neatly summed up in Figure 9. The comparatively poor performance of the workhorse of the chemical industry that is the jacketed stirred tank reactor is in stark contrast with the intensified systems.

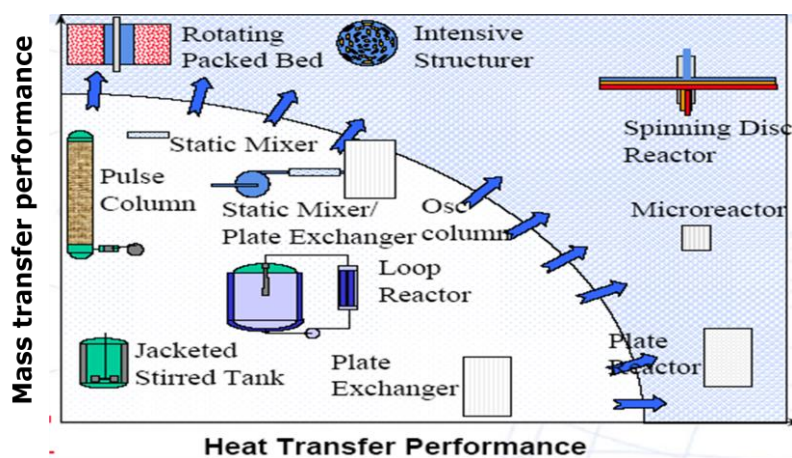


Figure 9. Heat and mass transfer performance of several intensified units compared to the jacketed stirred tank [67]

4.2. Integrating process steps

One of the most important intensification strategies is combining process steps together, thereby reducing the overall number in a process. This can have a range of benefits:

- i. Reduction in capital cost
- ii. Reduction in running cost

iii. Reduction in duty of downstream unit operations

One of the most familiar integrations is between reaction and separation. Examples include:

- **Membrane reactors:** membranes that are permeable to only one of the products are used to remove this product in situ, thereby overcoming equilibrium constraints. In this way membrane reactors have often been used to equilibrium reactions to completion.
- **Reactive extraction:** this is a separation process in which reaction is used to cause a product to move into a different phase. This is usually between two liquid phases, usually organic and aqueous, although it can be between solids and liquids. Again, this is often a way of overcoming equilibrium limitations.
- **Reactive distillation:** in this method the distillation column has a second function, as a reactor, thereby removing the reactor from the flowsheet. The greatest advantage lies not in reducing the number of unit operations in the plant, but in overcoming the equilibrium limitations of a process, by removing the product by distillation.
- **Supercritical operation:** a key advantage of performing reactions in supercritical carbon dioxide, for example, is that separation of the solvent (the CO₂) from the reaction mixture, becomes facile, and does not require a separate unit operation. Instead all that is required is that the pressure is released to the required degree for the reaction mixture to come out of solution.

There are many other examples: these are simply a selection to illustrate the range and possibilities for process improvement.

To focus on one example, reactive distillation (RD), in the 200 kte/yr Eastman process for methyl acetate production the following reaction takes place:



They successfully used reactive distillation to replace 11 distillation vessels (with associated condensers and reboilers), with just 3 RD vessels [68], resulting in significant running and capital cost savings (as distillation columns are amongst the most energy-intensive unit operations in any process). This intensification was successful, because the reaction fitted the requirements of RD. The most important of these was that one of the products (the methyl acetate itself) had a significantly lower boiling point (57 °C) than any other species present.

When integrating process steps, as with any other intensification, the effects on the whole process should be assessed. It may be that the effects on the process as a whole are minimal or even undesirable, or that the economics are not significantly affected by the new design. However, there is a wide range of process steps that can be combined, and with any process it is worth taking the time to assess the scope for intensification by integrating steps.

4.3. Moving from batch to continuous processing

Generally, continuous processing is more efficient than batch processing. Continuous reactors are usually smaller than equivalent batch for two reasons:

- 1. Greater Occupancy:** batch vessels are tied to the “batch cycle” meaning that, depending upon the process, much of the time can be spent on processes such as filling emptying, cleaning, heating and cooling, all of which reduce the space time yield of the vessel, as during all of these processes the reaction product is not being produced.

2. More Effective Mixing: the more efficient mixing observed in continuous reactors often leads to higher rates of reaction, which leads to smaller reactors (for the same production rate). This is only the case when the reaction is “mixing-limited”, but this often occurs at larger scales, because good mixing is difficult to achieve at scale, which limits the reaction rate. Batch stirred vessels “do not scale up well”, which essentially means that their performance at laboratory scale/pilot scale may not be a reliable guide to their performance at industrial scale. A particular problem is that energy densities that are easily achieved at laboratory scale are practically unattainable at the scale of 10s of m³, as the power scales up on the diameter of the impeller to the power 5. Continuous reactors often have better mixing simply because the length scales are smaller, the diameter of a tubular reactor being considerably less than that of an equivalent tank. Beyond this there may well be different mixing patterns in continuous reactors that are more effective than those of typical batch agitation.

The reduced scale and improved mixing patterns lead to a range of benefits of continuous operation such as:

- Improved control
- Better product quality
- Less hazardous operation, due to the reduced volumes of hazardous material
- Reduced capital costs: these are usually decreased due to the smaller size of the equipment, although this can in some instances be counteracted somewhat by costs associated with the mixing mechanisms or pumps.
- Reduced running costs: Better control of the temperatures in a reactor usually means that less energy is required to run the process, as the reaction mixture will not need to be overheated to allow for inhomogeneities. When reaction times are reduced by

conversion to continuous from batch, the reaction mixture has to be held at the correct temperature for a shorter period of time, which reduces the heating duty overall.

It should be noted that these are generalities, and as such they are not all true for all cases of batch-to-continuous conversion.

In practice, however, batch processing is often used, and there are a range of reasons for this:

1. Flexibility: batch vessels can be used for a range of different reactions, and indeed different process steps, whereas continuous equipment needs to be purpose-designed.
2. Batch integrity: in pharmaceutical processing, every operation undergone by a specific batch of reactants must be clearly defined, and this is achieved much more easily when there is a well-defined batch of material.
3. Small scale operation: at very small scales of production, the benefits of moving to continuous operation can be minimal. Continuous processing is usually one of the main economies of scale in chemicals manufacture.
4. High added value products: when the value of the product is high the processing costs can be immaterial, and other qualities of the process will become more important. Often the time-to-market will be critical, and in such cases, it is often perceived to be less risky to rely on known batch processes, or to scale up in batch, as the processes have been developed in the laboratory in batch.
5. Long residence times: when reaction times are more than a few minutes conventional reactors start to become increasingly impractical or expensive, if tight control of product specification (achieved by “plug flow”) is to be realised. Very few reactions that take

hours are performed continuously. This is the main “niche” of the Oscillatory Baffled Reactor (OBR): converting “long” batch processes to more efficient continuous processing.

5. Merits of PI Technologies

PI is associated with several business, process or environment related benefits, as summarized in Figure 10.

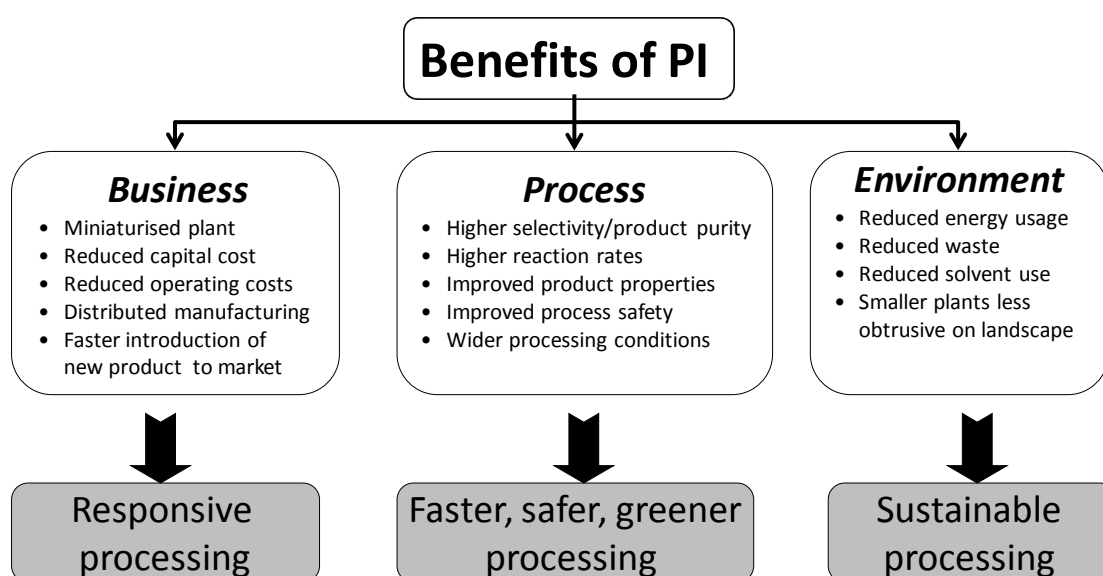


Figure 10. Business, process and environmental benefits of PI

5.1. Business

Cutting down the size of Main Plant Items such as reactors, heat exchangers and separators in a process not only entails a reduction in individual equipment cost but also, and to a more appreciable extent, huge cost savings are to be expected in installation costs as a result of reduced pipeworks, civil engineering and support structure to name but a few. Furthermore, it is envisaged that smaller sized plants would become more ‘mobile’ and hence be transported to the customer or to resources such as an oil field, for instance, for processing of flare gas into methanol. In this distributed manufacturing scenario, the logistics associated

with such small, portable plants will be greatly simplified. Another benefit, especially relevant to the pharmaceutical industry, is the ability to introduce new products to the market in a more rapid manner than is currently possible. With PI, a lab scale continuous reactor can become the manufacturing unit, if the throughput matches the desired production rate. This concept has been indeed been demonstrated in a feasibility study undertaken by SmithKline Beecham in the spinning disc reactor [69] where a lab scale spinning disc reactor could match the production capacity of a batch process giving 8 tonnes of product per annum. Cutting down on the process development stages can avoid lengthy delays as these stages alone can take up to 2-3 years when scaling up batch processes, which is a considerable amount of time in a product lifespan of 25 years before patent expiry.

5.2. Process

The process benefits of higher selectivity, faster reaction rates and better product properties have been explored in much detail under the various technologies considered in section 4 of this chapter.

Safety in the chemical industry is one of the highest priorities for Health & Safety regulatory bodies after a number of fatal incidents have occurred involving large inventories of hazardous materials and exothermic runaways in batch reactors such as the Bhopal disaster in 1984 [70]. Adopting the PI approach can substantially improve the intrinsic safety of a process by having a significantly reduced volume of potentially hazardous chemical at any time in a smaller intensified unit. In addition, one of the objectives of PI is to move away from batch processing to small continuous reactors, the latter giving more efficient overall operation especially in the case of hugely exothermic reactions whereby the heat can be removed continuously as it is being released. It is to be borne in mind though that by intensifying processes, for example by integrating process steps, the process is prone to

become more complex and overall safety considerations on the basis of a number of parameters need to be properly assessed, as highlighted in a recent study [71].

Another major advantage of PI technologies is that by virtue of their greatly improved heat and mass transfer they allow process conditions to be employed which would have previously been impossible with conventional technologies. This is what Hessel refers to as the ‘novel process windows’ offered by PI [72]. Several applications particularly related to microreactors [72-74] and spinning disc reactors [75, 76] where these more intense operating conditions such highly concentrated reaction systems, higher temperatures and higher pressures have been investigated are documented in the literature.

5.3. Environment

PI technologies have tremendously appealing environmental implications whereby a small, compact, highly intensified plant is more likely to be below the tree line, making it far less of an eyesore for the general public than the unsightly and massive steel works characterising our present chemical plants. Furthermore novel reactor designs based on the PI concept will enable clean technology to be practiced by enabling waste minimization at source. In other words, high selectivity operation in intensified reactors will reduce or eliminate altogether the formation of unwanted by-products which, if not removed from the effluent before discharge, can cause irreversible damage to the environment. High purity product and hence of improved quality will thus be obtained without incurring enormous downstream purification costs.

The improved energy efficiency foreseeable in intensified unit operations constitutes yet another highly attractive benefit of PI in a world where there is currently overwhelming concern over the ever growing demand on non-renewable energy resources. In order to address these concerns, there is a great and urgent need for the development of new process

technologies which will utilise energy in an efficient manner. In this respect, PI is a positive step in the right direction for the chemical industry. Large enhancements in heat and mass transfer, two of the most fundamental and frequently encountered operations in chemical engineering processes, can be achieved in intensified units, as has been described in earlier sections. Such improvements mean that process times and therefore the associated energy consumption can be dramatically reduced for a given operation. Furthermore, alternative energy sources such as microwave and light energy which can be targeted to desired process chemistries can result in less energy waste as is often encountered with conventional thermal energy sources. A case study highlighting the energy saving potential of a process intensified technology is presented in Chapter 15.

6. Challenges to Implementation of PI

Process intensification was first developed as a coherent philosophy by Ramshaw *et al.* in the 1970s so might be expected to have had more impact by now. There are manifold reasons for this lack of implementation in industry. Some are:

1. Perception of Risk: most chemical/pharmaceutical companies are (understandably) risk-averse, and for new technologies the risk is difficult to assess. There is, typically, a lack of data and a lack of previous case studies on which to base designs and operation, and therefore economic calculations. This results in a so-called “rush-to-be second” mentality. The lack of risk and cost data makes it difficult to assess the risk-reward balance and perform a reliable cost-benefit analysis for the process change.

2. Lack of “Champions” within Industry: for any given change in an organization, there is an associated individual with responsibility for it. When this change is substantial and technical, such as the change to a novel unit operation in a chemical plant, the “champion”

will have a lot of inertia to overcome, and must be dedicated and well-informed. The champion has to fully understand the technology and its benefits. It is the responsibility of technology developers to inform possible users of the benefits of their technology, but these changes cannot be effected without a substantial “pull” from the users.

3. Control/monitoring: the greatly enhanced rate of some intensified processes leads to problems in monitoring the process, as the response times of the instruments become significant in comparison to the process. This can be a significant added technical development step, and as such an obstacle to commercialisation. To address these issues, a large body of research work has been undertaken on control of more established PI technologies such as reactive distillation and dividing wall columns, as highlighted in a recent review of control aspects in PI [77]. Work is also currently on-going with newer developments such as the spinning disc reactor [78] to develop appropriate control strategies for these systems.

4. PI's Limitations: sometimes PI is not the solution. Any technology has a specific problem or set of problems that it can solve, and these need to be clearly identified and explained. The greatest danger of not doing so is that feasibility studies /pilot scale trials are performed that fail because the premise was incorrect (i.e. it was the wrong technology from the start), rather than any technical failing, and the technology itself is then labelled as inadequate. A number of technologies developed have been characterised as “solutions seeking problems”, and the problems will sometimes be found. However, sometimes the question that is asked is “can this technology be used for this application?”, when it should be “does this technology have realisable economic benefits for the process overall?”

Many of these challenges are familiar to anyone with a new technology of any sort to promote. The key issue is simply that the new process/technology must have a ***proveable and significant*** economic advantage. It must be borne in mind that an economic advantage is not achieved simply by reducing the size of a piece of process equipment (or, preferably the entire plant), economic advantages can be achieved in variety of ways:

- **Safer operation:** accidents are a cost. They should be factored into the economic evaluation of any project. Intensified technologies in general will significantly reduce the hazard, and therefore cost of any incident. More effective monitoring should help to reduce the rate of incidence of accidents.
- **Product quality:** improved product quality reduces the load on downstream separations, possibly allowing them to be reduced in size, or alternative technologies to be used, or process steps to be removed altogether, thereby reducing capital and running costs in other areas of the plant. The higher product quality produced in intensified processes due to the enhanced effectiveness of heat and mass transfer reduces the amount of waste produced, thereby lessening the waste to be disposed. In some industries where the wastes are highly toxic, for instance, disposal and/or treatment is one of the largest running costs. This will become more of a driver in the future, as environmental legislation becomes more stringent worldwide.
- **Reduced running costs:** the greater efficiency of intensified plants reduces energy costs. In a fossil fuel-based energy supply system, as most currently are, this equates directly to reduced CO₂ emissions. Such issues should become increasingly important as carbon trading becomes more important.

7. Concluding Remarks and Future prospects

As a revolutionary approach to chemical processing, PI has gained remarkable momentum in the last two decades. It has been described more than a decade ago as “the key to survival of the fittest in international competition” [79]. It is not surprising therefore, given the intensely competitive environment businesses in the chemical industry operate in, that great strides have been made in research and development of PI technologies over the last few years.

With greater emphasis being placed on sustainable development nowadays, process intensification can be an important element in making the future chemical and pharmaceutical industries greener. This is because it can be used to greatly reduce the size of many unit operations, and in so doing reduce not only the energy required to run them but also the surrounding infrastructure and the energy and materials used in their manufacture. In an intensified environment, higher selectivities and yield can result in improved atom efficiency and waste minimisation at source, leading to reduced downstream processing. There is no doubt that process intensification is here to stay and will continue to play a major role in the development and growth of chemical and processing industry in the years to come. Whether or not large plants can be considered as ‘dinosaurs’ that will be extinct within the next 50 years or so remains to be seen, but it is a vision that is not altogether unrealistic. But there are still many challenges ahead which will require a concerted effort from the academic community as well as industry partners before PI is more widely implemented and practised. It is also worthy of note here that how successful PI is in its application to a given process is very much dependent on the characteristics of the process. An inherently slow process will not be made any faster by intensifying heat and mass transfer characteristics. Thus, it is important to clearly understand the fundamentals of a process before any attempt at intensifying it is made. There are a few other process-related limitations to general application of PI such as:

(i) some processes operate at too small a scale for process intensification to be economically worthwhile;

(ii) various plants require a flexibility of operation that is difficult to achieve in continuous plant, which is usually purpose-designed

However, in spite of these application limitations and challenges, the potential for process intensification remains largely untapped, but the economic rewards for those companies that do introduce PI are likely to be substantial - and this will have environmental benefits for a more sustainable future for generations to come.

Nomenclature

A	Constant
A _s	Heat transfer surface area (m ²)
C ₁ , C ₂	Constants in Nusselt number and Sherwood number correlations respectively
C _D	Coefficient of discharge of impeller (-)
C _p	Specific heat capacity (J.kg ⁻¹ .K ⁻¹)
D	Pipe or channel diameter (m)
D	Diffusivity (m ² .s ⁻¹)
D _i	Impeller diameter (m)
D _l	Liquid diffusivity (m ² .s ⁻¹)
D _{turb}	Turbulent diffusivity (m ² .s ⁻¹)
h	Heat transfer coefficient (W.m ⁻² .K ⁻¹)
k	Thermal conductivity (W.m ⁻¹ .K ⁻¹)
k _m	Mass transfer coefficient (m.s ⁻¹)
l	Characteristic length for diffusion (m)
L	Characteristic length scale of large eddies (at macroscale) (m)

L_c	Integral length scale of turbulence (m)
L_D	Characteristic length scale for turbulent dispersion (m)
N	Impeller rotational speed (Hz)
Nu	Nusselt number ($=hD/k$)
Q	Heat transfer rate (W)
Q_c	Impeller pumping capacity ($m^3.s^{-1}$)
Q_f	Flowrate of injected fluid ($m^3.s^{-1}$)
r_{pipe}	Radius of pipe for feed introduction (m)
Re	Reynolds number ($=\rho uD/\mu$)
Sc	Schmidt number ($=\nu/D$ or ν/D_l)
Sh	Sherwood number ($=k_m.l/D$)
t_D	Characteristic diffusion time (s)
ΔT_{lm}	Logarithmic mean temperature difference (K)
u	Average velocity ($m.s^{-1}$)
U	Overall heat transfer ($W.m^{-2}.K^{-1}$)
V	Vessel volume (m^3)

Greek letters

ε	Turbulent energy dissipation rate ($W.kg^{-1}$)
η_K	Kolgomorov length scale (m)
η_B	Batchelor length scale (m)
ν	Kinematic viscosity of fluid ($m^2.s^{-1}$)
τ_c	Circulation time (s)
τ_D	Characteristic mesomixing timescale for turbulent dispersion (s)

τ_{DS}	Characteristic micromixing time for shear deformation and diffusion at Batchelor scale (s)
τ_e	Characteristic micromixing time for engulfment (s)
τ_{mac}	Macromixing time (s)
τ_s	Inertial-convective mesomixing time (s)
μ	Viscosity (Pa.s)
ρ	Density (kg.m ⁻³)

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